

Please replace the paragraph beginning on page 1, line 10 with the following:

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According to the conventional process for producing polymers of polycondensation series of polyethylene terephthalate, etc., raw materials terephthalic acid and ethylene glycol are fed to a mixing tank in an appropriate proportion for esterification and the mixture is then fed to esterification reactor tanks through a pump. In the esterification step, 2 or 3 stirring tanks each provided with stirring vanes are provided in series and a water effluent as a by-product is separated in a distillation column. A prepolymerization step follows, where a plurality of vertical stirring tanks or horizontal stirring tanks are provided, and a final polymerization step follows, where a horizontal stirring tank is provided. The tanks of these polymerization steps are each provided with a condenser for removing ethylene glycol as a by-product, and are operated in a reduced pressure atmosphere. In the conventional process for producing polyester there are 4 to 6 reactor tanks, each of which is provided with stirring vanes and its power source and further provided with a distillation column or a condenser for separating and removing the by-product. Furthermore, the polymerization steps are operated in a reduced pressure atmosphere, and thus the necessary additional vacuum means for generation of a reduced pressure is required. That is, the operation of the apparatus requires a higher maintenance cost and a higher apparatus cost. One example of the related art is disclosed in JP-A 7-207009, where each tank is provided with stirring vanes and is vacuum-controlled, and thus its running or operation is complicated. That is, improved apparatus and process with a higher efficiency

throughout the apparatus and economical operation based on energy saving  
throughout the apparatus have been so far desired for the production of high  
molecular weight polyester.

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Please replace the paragraphs beginning on page 3, line 12 and ending on page 8,  
line 26 with the following:

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According to a first aspect of the present invention, a process for continuously  
producing polyester is provided, which comprises a first step of reacting an aromatic  
dicarboxylic acid or its derivative with a glycol in a first reactor, thereby producing an  
oligoester or a polyester having an average degree of polymerization of 3 to 7; a  
second step of polycondensing the oligoester or polyester of the first step in a  
second reactor, thereby producing a low molecular weight polyester having an  
average degree of polymerization of 20 to 40; and a third step of further  
polycondensing the low molecular weight polyester of the second step to an average  
degree of polymerization of 90 to 180 in a third reactor, thereby producing a high  
molecular weight polyester, wherein at least one of the first reactor and the second  
reactor is free from an external power source stirring means; or wherein a reactor is  
used as the third reactor, which comprises a horizontal cylindrical vessel having an  
inlet for the low molecular weight polyester of the second step at the lower part at  
one end thereof and an outlet for the high molecular weight polyester at the lower  
part at the other end thereof in the longitudinal direction thereof, an outlet for volatile  
matters at the upper part thereof, and a stirring rotor rotatable in the vicinity of the  
inner periphery of the vessel in the longitudinal direction of and within the vessel, the

A3 stirring rotor within the vessel comprising a plurality of stirring blocks according to the viscosity level of the polyester, and the stirring rotor having stirring vanes without any rotating shaft at the center of the stirring rotor; more preferably wherein a mixture of the aromatic dicarboxylic acid or its derivative and the glycol in a molar ratio of 1:1.05 to 1:2.0 of the former to the latter is fed to the first reactor kept at a temperature of 240° to 285°C under a pressure of the atmospheric pressure to  $3 \times 10^6$  Pa; the oligoester or polyester from the first reactor is fed to the second reactor kept at a temperature of 250° to 290°C under a pressure of the atmospheric pressure to 133 Pa; and the low molecular weight polyester from the second reactor is fed to the third reactor kept at a temperature of 270° to 290°C under a pressure of 200 to 13.3 pa; more preferably wherein the stirring rotor is rotated at 0.5 to 10 rpm; and more preferably wherein total reaction time throughout the first reactor, the second reactor and the third reactor is 4 to 8 hours.

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Pl According to a second aspect of the present invention, an apparatus for continuously producing polyester is provided, which comprises a first reactor for reacting an aromatic dicarboxylic acid or its derivative with a glycol, thereby producing an oligoester or polyester having an average degree of polymerization of 3 to 7; a second reactor for polycondensing the oligoester or polyester from the first reactor, thereby producing a low molecular weight polyester having an average degree of polymerization of 20 to 40; and a third reactor for further polycondensing the low molecular weight polyester from the second reactor to an average degree of polymerization of 90 to 180, thereby producing a high molecular weight polyester, wherein at least one of the first reactor and the second reactor is free from an

external power source stirring means; or wherein the third reactor comprises a horizontal cylindrical vessel having an inlet for the low molecular weight polyester from the second reactor at the lower part at one end thereof and an outlet for the high molecular weight polyester at the lower part at the other end thereof in the longitudinal direction thereof, an outlet for volatile matters at the upper part thereof, and a stirring rotor rotatable in the vicinity of the inner periphery of the vessel in the longitudinal direction of and within the vessel, the stirring rotor within the vessel comprising a plurality of stirring blocks according to the viscosity level of polyester feed and the stirring rotor having stirring vanes without any rotating shaft at the center of the stirring rotor; more preferably wherein the first reactor is conduct the reaction at a temperature of 240° to 285°C under a pressure of the atmospheric pressure to  $3 \times 10^6$  Pa; the second reactor is to conduct the polycondensation at a temperature of 250° to 290°C under a pressure of the atmospheric pressure to 133 Pa, and the third reactor is to conduct the further polycondensation at a temperature of 270° to 290°C under a pressure of 200 to 13.3 Pa.

The above-mentioned objects of the present invention can be attained by using three single reactors each for a esterification step, a prepolymerization step and a final polymerization step, where single reactor without any external power supply are used for at least one of the esterification step and the prepolymerization step, and a stirring power-requiring reactor is used for the final polymerization step. The first reactor for the esterification step according to the present invention is, for example, a natural circulation type evaporator comprising a vertical cylindrical vessel provided with an inlet and an outlet for a raw material feed such as a mixture an

AB aromatic dicarboxylic acid or its derivative and a glycol and an oligoester or polyester at the bottom of the vessel and further with a vapor pipe for discharging the vapors at the top of the vessel; a heating medium jacket covering the outside of the vessel; and a shell-and-tube type heat exchanger within the vessel, the tube outside of the heat exchanger being heated by a heating medium, while allowing the raw material feed to rise through the tubes to make the external power unnecessary. It is desirable for the natural circulation type evaporator to make the average speed of the liquid feed flowing downwards by natural convection between the inner wall of the cylindrical vessel and the outer wall of the shell of the shell-and-tube type heat-exchanger smaller than the average speed of the liquid feed rising through the tubes of the shell-and-tube type heat exchanger and also to provide an entrance space at the bottom of the shell-and-tube type heat exchanger so as to allow the internally circulating raw materials to uniformly enter into the tubes.

Me For the second reactor for the prepolymerization step is used an apparatus comprising, for example, a substantially vertical cylindrical vessel provided with an inlet for an oligopolyester or polyester feed from the first reactor and an outlet for a low molecular weight polyester feed on the lower end side and at the bottom at the center of the vessel, respectively, in the longitudinal direction of the vessel, and further with an outlet for volatile matters at the top of the vessel, and covered with a heating medium jacket on the outside of the vessel, thereby to make any external power unnecessary. The second reactor may be further provided with a heat-exchanging section at the lower part within the vessel and a residence section with helical baffle plates at the intermediate part within the vessel so as to retain the

AB liquid feed and transfer the polyester feed successively from a lower stage to a upper stage, further with a space for gas-liquid separation at the upper part within the vessel and a down pipe in the vertical direction at the center within the vessel so as to allow the polyester feed to flow downwards as a thin film.

AM For the third reactor for the final polymerization step is used an apparatus comprising, for example, a horizontal cylindrical vessel provided with an inlet for the low molecular weight polyester feed from the second reactor at the lower part at one end and an outlet for a high molecular weight polyester at the lower part at the other end, respectively, of the vessel in the longitudinal direction thereof, and further with an outlet for volatile matters at the top of the vessel and a stirring rotor, which is extended in the longitudinal direction of the vessel and rotates in the vicinity of the inner periphery of the vessel, the stirring rotor within the vessel being divided into a plurality of stirring blocks according to the viscosity level of the polyester feed, and the stirring rotor being without a rotating shaft but with stirring vanes at the center part of the stirring rotor.

AM Please replace the paragraph beginning on page 9, line 15 with the following:

Fig. 6 is a cross-sectional view along the line B-B of Fig. 4.

AM Please replace the section heading and the paragraph beginning on page 10, line 11 with the following:

#### BEST MODES FOR CARRYING OUT THE INVENTION

AM Fig. 1 shows one embodiment of the present invention, i.e. a flow diagram

Ag showing an arrangement of reactors for a process for continuously producing polyethylene terephthalate. Economically very advantageous industrious process for producing polyester is a direct esterification process, which has been recently much employed. In Fig. 1, numeral 31 shows a feed mixing tank for mixing and stirring TPA (terephthalic acid) as an aromatic dicarboxylic acid and EG (ethylene glycol) as a glycol in a molar ratio of 1:1.05 to 1:2.0 of the former to the latter both being raw materials for producing polyethylene terephthalate. At this esterification stage of the process, a polymerization catalyst or additives such a stabilizer, a color tone regulator, etc. can be used together, when required.

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Please replace the paragraph beginning on page 11, line 15 and ending on page 15, line 12 with the following:

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Ad Development of new catalysts has been so far attempted to overcome these problems. Antimony compounds now used most in the industry, particularly antimony trioxide, are distinguished in the cost and effects. However, discoloring of the resulting polyester polymers is inevitable even by using these catalysts. To overcome this disadvantage, a phosphorus-based stabilizer (for example, trimethyl phosphate and triphenyl phosphate) is used together as a stabilizer. In another attempt, an inlet position of the polymerization catalyst or the stabilizer is elaborated to stabilize the quality. In the ordinary process, it is preferable to use 200 to 400 ppm of the catalyst and 50 to 200 ppm of the stabilizer.

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Am The raw materials so mixed are led to an esterification reactor 33 as a raw material feed through a feed line 32 supplying the raw material feed thereto. The

esterification reactor (first reactor) 33 is provided with a heating medium jacket structure (not shown in the drawing) on the outer periphery of the reactor to keep the raw material feed at a reaction temperature and also provided with a shell-and-tube heat exchange 34 as dipped in the raw material feed within the first reactor as a means of heating the raw material feed from an external heat source and circulate the raw material feed within the first reactor by natural circulation, thereby proceeding with the reaction. Type of most desirably first reactor is such a calandria type that the esterification reaction is to be carried out by naturally circulating the raw material feed within the reactor on the basis of utilization of the evaporating action of byproducts formed by the reaction. This type of reactor needs no external power source for stirring, resulting in such advantages as a simple reactor structure, unnecessary for a stirring shaft-sealing device and a low reactor-manufacturing cost. One example of such a desirable first reactor is shown in Fig. 2. Fig. 2 shows one embodiment of the first reactor. Raw material feed 52 flows into a vertical evaporator 51 through an inlet 53 provided at the bottom of the evaporator 51 and is heated while passing through a plurality of heat transfer tubes in a shell-and-tube type heat exchange 54 to go upwards by natural convection, where a portion of low boiling point components in the raw material feed 52 is evaporated, and discharged from a vapor pipe 55 to the system outside. The remaining raw material feed 52 flows downwards between the inner wall of the evaporator 51 and the outer wall of the shell of the shell-and-tube type heat exchange 54 by natural convection and enters into a cylindrical entrance space 56 provided at the bottom of the shell-and-tube type heat exchange 54, where the flow of the raw material feed can be rectified to a less



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turbulent state, and the average flow speed through the tubes of the shell-and-tube heat exchanger 54 is made higher than the average flow speed of the raw material feed flowing downwards by natural convection between the inner wall of the evaporator and the outer wall of the shell. Thus, the raw material feed can enter into a plurality of heat transfer tubes with a more uniform speed distribution and uniformly heated again to repeat circulation by natural convection. In the course of circulation, the low boiling point components are evaporated and after an appropriate convection time, the resulting oligoester or polyester 59 is led to the system outside from an outlet 60. To generate a smooth, accelerated flow, the flow passage area of the cylindrical entrance space must be designed larger than the total flow passage area of heat transfer tubes and also the passage area of double pipe section formed between the inner wall of the evaporator 51 and the outer wall of the shell of the shell-and-tube type heat exchanger 54 is made larger than the flow passage area of the entrance space. Numeral 57 shows an inlet of a heating medium and 58 shows an outlet of the heating medium. The outer periphery of the evaporator 51 is covered with a heat insulator or a heating medium jacket (not shown in the drawing). Thus, in the evaporator of this embodiment, uniform flow speed distribution in the axial direction of the heat exchanger can assure uniform evaporation or esterification reaction of the raw material feed, and a better product quality can be effectively obtained in a short residence time. In case that the raw material feed 52 is a mixture of solid particles and a liquid (which will be hereinafter referred to as slurry), the raw material feed 52, which undergoes natural circulation, flows into the cylindrical entrance space 56 provided at the bottom of the shell-and-

AN1 tube type heat exchanger 54, and by more smooth rising along a conical member 62, the solid particles are never precipitated at the bottom. That is, when the liquid is a slurry, solid particles contained in the slurry can be prevented from precipitation by providing at the bottom of the evaporator the conical member for rising the raw material feed to undergo internal circulation. The conical member may have some curvature.

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AN2 Thus, the evaporator of this embodiment is effective and more suitable for natural circulation of a slurry, and can produce a reliable polyester product of good quality. The present invention is not limited to this type of evaporator and use of an evaporator with stirring vanes, is not objectionable for process reasons.

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AN3 In the first reactor, water formed by the reaction is in a vapor form, and forms a gas phase 65 together with vaporized EG. Recommendable reaction conditions for the first reactor are desirably a temperature from 240° to 285°C and a pressure of the atmospheric pressure to  $3 \times 10^6$  Pa. The gases in the gas phase 65 are separated into water and EG by a rectification column (not shown in the drawing), and water is discharged to the system outside, whereas EC is returned to the system.

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Please replace the paragraphs beginning on page 15, line 20 and ending on page 18, line 28 with the following:

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AN4 In Fig. 1, the raw material feed, when retained in the esterification reactor 33 (first reactor) for a predetermined reaction time reaches a predetermined rate of esterification to produce an oligoester or polyester having an average degree of

polymerization of 3 to 7 and then the resulting oligoester or polyester is supplied to an initial polymerization reactor (second reactor) 37 through a connecting pipe 36, where the oligoester or polyester feed is heated to a predetermined reaction temperature by a heat exchanger 38 to undergo polycondensation reaction, thereby increasing the average degree of polymerization. The reaction conditions are a temperature of 250° to 290°C and a pressure of the atmospheric pressure to 133 Pa, preferably 266 Pa to 133 Pa, whereby the polycondensation reaction is made to proceed to produce a low molecular weight polyester having an average degree of polymerization of 20 to 40. The initial polymerization reactor shown in this embodiment is a reactor without stirring vanes, but the present invention is not limited to such a reactor. The initial polymerization stage is such that kinetics of polymerization reaction is rate-determining and thus the reaction can proceed smoothly only by supplying the necessary quantity of heat for the polycondensation reaction fully. Thus, there is no need for the oligoester or polyester feed to unnecessitate the stirring action by stirring vanes and it is only necessary to discharge EG formed by the polycondensation to the system outside. Most suitable second reactor for such an operation is desirably an apparatus shown in Fig. 3.

In Fig. 3, numeral 71 is a vertically long, cylindrical vessel whose outer periphery is covered by a heating medium jacket 72 having an inlet 80 for a heating medium and an outlet 81 for the heating medium, and a down pipe 73 having an open top end is provided in the longitudinal direction at the center of the vessel 71. A plurality of heat transfer tubes 74 are provided at the lower part within the vessel 71 in parallel to the down pipe 73. A plurality of helical baffle plates 75 are provided on

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given different levels on the outer periphery of the down pipe 73 above the heat transfer tube 74. The baffle plates 75 each have a clearance 83 between its edge and the inner wall of the vessel 71 to allow volatile matters to pass therethrough and partitions the interior of the vessel 71 in the vertical direction to form a plurality of residence compartments 84. There is a space 76 for separating the liquid feed from volatile matters at the upper part of the vessel 71, that is, above the down pipe 73 and the uppermost baffle plate 75C. A plurality of tapered liquid receivers 88 are provided on given different levels on the inner wall of the down pipe 73, through which the oligoester or polyester feed flows downwards as a thin film. The oligoester or polyester feed flowing downwards through the down pipe 73 can be retained once in the individual liquid receivers and then successively move downwards, and thus the polycondensation reaction can proceed, while minimizing the oligopolymer or polyester feed short path and effectively separating the vapors of volatile matters.

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In the second reactor, the oligoester or polyester feed continuously supplied through an inlet nozzle 77 enters into heat transfer tubes 74 and rises therethrough while being heated and reaches the lowermost residence chamber 84A. While the feed slowly rises through the residence compartment 84A, polycondensation reaction proceeds and the resulting volatile matters such as ethylene glycol, etc. move upwards through the clearances 83 at the outer edge of the baffle plate 75. On the other hand, the feed rises along the helical configuration of the baffle plate 75 in a whirling stream in the residence compartment 84A and enters into next upper residence compartment 84B. Since the feed can move smoothly in a whirling stream into the next upper residence compartment 84B, the feed can rise successively

through other upper residence compartment without causing any backflow, and thus the polycondensation reaction proceeds effectively.

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The feed, which has reached the uppermost 15 residence compartment 84C, flows over the top edge 82 of the down pipe 73 and flows downwards along the inner periphery of the down pipe 73 as a thin film, while separating vapors of volatile matters resulting from the reaction, and thus the polycondensation reaction can further proceed. The feed at an advanced stage of polycondensation reaction, which has been separated from vapors of volatile matters resulting from the reaction, is discharged to the system outside through an outlet nozzle 78, while the resulting volatile matters are separated from entrainments of the feed (polyester) in an upper space 76 in the vessel 71 and discharged to the system outside through an outlet nozzle 79 for the volatile matters.

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Please replace the paragraph beginning on page 19, line 20 with the following:

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In Fig. 1, the feed retained in the initial polymerization reactor (second reactor) 37 for a predetermined reaction time is supplied to a final polymerizer (third reactor) 41 through a connecting pipe 40. In the final polymerizer the polycondensation reaction further proceeds under a good surface renewal action of stirring vanes 42 without a stirring shaft at the center to elevate the degree of polymerization, whereby a high molecular weight polyester having an average degree of polymerization of 90 to 180 is produced. Suitable final polymerizer (third reactor) is shown in Figs. 4 and 15, which have most distinguished surface renewal performance and power consumption characteristics. Owing to a broad viscosity

range of the liquid feed, the conventional final polymerizer is divided into two apparatuses to conduct the final polymerization step, whereas in the present invention the step can be carried out in a single apparatus, resulting in large reduction in the apparatus cost.

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Please replace the paragraphs beginning on page 20, line 18 and ending on page 24, line 4 with the following:

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In Fig. 4, numeral 1 shows a horizontally long cylindrical vessel, whose outer periphery is covered with a heating medium jacket (not shown in the drawing), and at whose both ends in the longitudinal direction are provided shafts 3a and 3b for rotor support, respectively. A stirring rotor 4 is fixed to and supported between the shafts 3a and 3b for the rotor support, and the rotor shaft 3a is connected to a driving means (not shown in Fig. 4, but shown as 44 in Fig. 1), and has rotor support members 2a and 2b connected to connecting support rods 5a, 5b, 5c and 5d, as shown in Figs. 4 and 5 (the number of connecting support rods depends on the size of the stirring rotor 4 but is 4 in this embodiment) at both ends of the stirring rotor 4. The stirring rotor 4 has a plurality of stirring blocks between the support members (end disks) 2a and 2b. The support member 2a is a member for a low viscosity polyester feed and the support member 2b for a high viscosity polyester feed. The support member 2b is smaller in the outer diameter than the stirring rotor 4 and has scraping vanes 13a and 13b on the vessel end-facing side to forward the high molecular weight polyester staying on the inner wall of the vessel 1 towards the

product outlet of the vessel 1 by rotation of the stirring rotor 4. Its detail structure is shown in Fig. 14, which is a cross-sectional view along the line E-E of Fig. 4. In the low viscosity zone of the stirring rotor 4 in the vicinity of an inlet nozzle 11 is provided a plurality of low viscosity stirring blocks, each of which comprises a pair of hollow disks 8 with buckets formed by scraping vanes 6a and 6b and thin hollow disks 7a provided between a pair of the hollow disks 8 and being subject to the low molecular weight polyester feed pouring from the buckets (its detailed structure will be described referring to Figs. 5, 9 and 10). In the intermediate viscosity zone is provided a plurality of intermediate viscosity stirring blocks, each of which comprises a pair of hollow disks 8, a plurality of thin hollow disks 7b having the same outer diameter at equal distances between the pair of hollow disks 8 and a plurality of scraping vanes 6c provided radially at the outer peripheral side of these disks (its detailed structure will be described, referring to Figs. 6, 7, 11 and 12). Furthermore, a high viscosity stirring block, which comprises a plurality of wheel-type disks 9 provided at appropriate distances from one another and scraping vanes 10 provided at the outer peripheral side of the wheel-like disks 9, is provided at the outlet side of the vessel 1 (its detailed structure will be described, referring to Figs. 3 and 13). An outlet nozzle 12 for discharging the high molecular weight polyester product (final polymers) is provided at the lower part at the other end of the vessel 1, and an outlet nozzle 14 for the volatile matters is provided on the upper part of the vessel 1 and is connected to a condenser and a vacuum means through a piping (not shown in the drawings).

In the final polymerizer as shown in Fig. 4, a low viscosity low molecular

weight polyester (prepolymer) feed having a low degree of polymerization, e.g. an average degree of polymerization of 20 to 40, continuously fed through the inlet nozzle 11 is at first stirred in a plurality of the low viscosity stirring blocks by buckets on the hollow disks shown in Fig. 5. At that time, the feed has such a low viscosity as a few Pas to a few tens Pas. In the low viscosity stirring blocks, one pair of the scraping vanes 6a and 6b forms one bucket at the outer peripheral side of the hollow disk 8, as shown in Fig. 5, so as to scoop up the feed into the bucket by rotation as shown in Fig. 9. That is, Figs. 9 and 10 schematically show feed-flowing states. At the bottom of each bucket of the scraping vanes 6a and 6b is formed a small hole or clearance 6. With rotation of the stirring rotor, low viscosity feed 91 is scooped up by the bucket, as shown by numeral 100 in Fig. 9, and the buckets are further inclined downwards by further rotation to allow the feed to start to leave the inwardly, and at the same time to start to leak out outwardly little by little from the bucket through the small hole or clearance, as shown by numeral 102 in Fig. 9, to form inward and outward liquid films 101 and 102, respectively, from the bucket. The feed 101 flowing downwards inwardly from the bucket pours onto the thin hollow disk 7a provided in the vicinity of the tip end at the inward side of the buckets, as shown by numeral 103 in Fig. 10, to form liquid films on the surface of each of the thin hollow disk 7a and between the adjacent thin hollow disks 7a at the same time, thereby producing a larger evaporation surface area. These actions are repeated by each rotation of buckets to assure a sufficient evaporation surface area and a better surface renewal. Satisfactory better performance can be obtained in the low viscosity stirring blocks even at a low revolution per minutes such as 0.5 rpm to a



A20  
few rpm (not more than 10 rpm), and effective reduction of stirring power consumption can be obtained. By-products evaporated from the feed pass through center holes 20a of the hollow disks 8 and center holes of thin hollow disks 7a and are given off through the outlet nozzle 14.

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Please replace the paragraphs beginning on page 24, line 25 and ending on page 28, line 24 with the following:

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A21  
The feed 92 having an increased viscosity such as a few tens Pas is brought upwards by the scraping vanes 6c by rotation, and flows downwards by gradual inclination of the scraping vanes 6c by further rotation thereof to form liquid films 104, as shown in Fig. 11. The liquid films 104 flow down onto the connecting support rods 5a of the stirring rotor 4 and are retained there for a long time as stuck thereto and suspended downwards therefrom, and furthermore the feed so brought upwards are suspended downwards over the center hole 20a of the hollow disk 8 to form a liquid film 105, as shown in fig. 11. A liquid film 107 is likewise formed over the center hole 20a of the thin hollow disk 7b and also the feed is suspended down over the small circular holes 20b of the thin hollow disk 7b to form liquid films 106, as shown in Fig. 12. The feed is formed into such liquid films and can have much increased degree of polymerization and viscosity due to much increased evaporation surface area and surface renewal.

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A22  
When the viscosity of the feed reaches a few hundred Pas, the feed is treated in a high viscosity stirring block. The high viscosity stirring block comprises a plurality of wheel type disks 9 having scraping vanes 10 (Fig. 4) at the outer peripheral side

A00  
thereof, as shown in Fig. 8. The wheel type disks 9 are connected to one another at predetermined distances by the connecting support rods 5a, 5b, 5c and 5d in the horizontal direction. The scraping vanes 10 comprise front side vanes 10a and back side scraping vanes 10b alternately arranged on one wheel type disk 9 and on another adjacent wheel type disk 9, respectively, so that the total length in the horizontal direction of the scraping vanes on all the wheel type disks 9 is such that the track of one scraping vane 10a on one wheel type disk can overlap that of one scraping vane 10b on another wheel type disk, when rotated, to scrape the corresponding entire inner wall surface of the vessel 1.

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A23  
As shown in Fig. 13, the feed 93 having a viscosity of a few hundred Pas is brought upwards by the scraping vanes 10a by rotation of the stirring rotor 4, and the feed so brought upwards flows downwards by rotation of the wheel type disk 9 and suspended to form a liquid film 108, and furthermore a liquid film 109 is formed also over the hollow parts of the wheel type disk 9, thereby creating complicated surface shapes of liquid films. When the viscosity of the liquid feed is further increased to reach a few thousand Pas, the amount of the feed so brought upwards by the scraping vanes 10a is increased. When the revolution per minute of stirring rotor 4 is increased in that state, such a retained feed-circulating phenomenon as to bring upwards the feed retained by the scraping vanes 10a again without leaving the vanes occurs. Thus, the stirring rotor 4 must be rotated at not more than 10 rpm. When the viscosity of the liquid feed becomes higher, the optimum rotation must be made slower. According to the inventors' test, the optimum rotation is in a range of 0.5 to 6 rpm. Stirring and surface renewal actions are repeated, as described above,

A23  
to promote polycondensation reaction. Volatile matters formed by the reaction pass through the hollow parts, center holes and small holes of the hollow disks successively and move in the longitudinal direction through the vessel 1 and are discharged to the system outside through the outlet nozzle 14 for the volatile matters. The resulting final high molecular weight polyester having a high degree of polymerization such as an average degree of polymerization of 90 to 180 and a high viscosity such as a few thousand Pas so obtained is discharged to the system outside through the outlet nozzle 12. At that time the final polyester having such a high viscosity is liable to stay in the region above the outlet nozzle 12, but the outer diameter of the support member 2b of the stirring rotor 4 is smaller than the outer diameter of the stirring rotor 4, the final polyester never stays on the support member 2b. Furthermore, the scraping vanes 13a and 13b are provided on the support member 2b on the vessel inner end wall-facing side, as shown in Figs. 4 and 14, to forward the final polyester therefrom towards the outlet 12 of the vessel 1, and thus the inner end wall surface of the vessel is kept always in a substantially complete self-cleaned state to prevent the final polyester from the deposition and staying thereon.

A24  
In the polycondensation of polyethylene terephthalate in such a final polymerizer, a prepolymer feed from the second reactor is continuously supplied into the final polymerizer (third reactor) through the inlet nozzle 11, stirred by the stirring rotor 4, while renewing the prepolymer feed surfaces, and evaporating and removing volatile matters resulting from the polycondensation reaction such as ethylene glycol, etc. and as a result of the polycondensation reaction final polyester of a higher

viscosity can be obtained. Volatile matters separated during the polycondensation reaction, for example, ethylene glycol, etc. are discharged to the system outside through the outlet nozzle 14. Operating conditions for the final polymerizer are a temperature of 260° to 300°C, preferably 270° to 290°C, a pressure of 10,000 to 10 Pa, preferably 200 to 13.3 Pa and a stirring rotor revolution of 1 to 10 rpm. The resulting final polymers are discharged to the system outside through the outlet nozzle 12. The interior of the vessel 1 is stirred and kept always in a substantially complete self-cleaned state during the polycondensation reaction and the feed is subject to better surface renewal, and thus final polyester of good quality can be efficiently obtained without any deterioration due to the staying in the vessel 1. This type of the present final polymerizer can be likewise applied to continuous bulk polymerization of resins of polycondensation series such as polyethylene naphthalate, polyamide, polycarbonate, etc.

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Please ~~replace~~ the paragraph beginning on page 29, line 14 and ending on page 29, line 26 with the following:  

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*A25*  
In the above-mentioned apparatus structure for producing polyethylene terephthalate, number of reactors is decreased, as compared with the conventional apparatus structure and thus the apparatus cost can be much reduced in the present invention.

*A26*  

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Not only can the number of accessory distillation or rectification columns and condensers be decreased due to the decreased number of reactors but also pipings connecting thereto, instrumentation or control devices and valves can be largely